U.S. DEPARTMENT OF COMMERCE
National Technical Information Service
PB-295 011

Flue Gas Desulfurization Pilot Study. Phase I Survey of Major Installations. Appendix 95-J. Sodium Sulfite Scrubbing Flue Gas Desulfurization Process

NATO Committee on the Challenges of Modern Society, Brussels (Belgium)

Jan 79

PB 295011

COMMITTEE ON THE CHALLENGES
OF MODERN SOCIETY

Flue Gas Desulfurization Pilot Study Phase I — Survey of Major Installations - Sodium Sulfite Scrubbing Flue Gas Desulfurization Process

Appendix 95-J

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U. S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

NORTH ATLANTIC TREATY ORGANIZATION

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. 95-J	2.
3. RECIPIENT'S ACCESSION NO. PB295011	
4. TITLE AND SUBTITLE Flue Gas Desulfurization Pilot Study Phase I - Survey of Major Installations - Sodium Sulfite Scrubbing Flue Gas Desulfurization Process	5. REPORT DATE January 1979
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) United States	8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS PEDCo Environmental Protection Agency Cincinnati, Ohio	10. PROGRAM ELEMENT NO.
	11. CONTRACT/GRANT NO.
12. SPONSORING AGENCY NAME AND ADDRESS U.S. Environmental Protection Agency Office of Research and Development (RD-681) 401 M Street, S.W. Washington, D.C. 20460	13. TYPE OF REPORT AND PERIOD COVERED
	14. SPONSORING AGENCY CODE EPA/600/00

15. SUPPLEMENTARY NOTES This report is one of a series on the status of flue gas desulfurization processes in the NATO countries and Japan. The United States, the Federal Republic of Germany, and Norway have acted as lead countries in the gathering of information for and preparation of the reports.

16. ~~ABSTRACT~~

Abstract: The Wellman-Lord flue gas desulfurization (FGD) process is currently the most recognized regenerable system using soluble sodium sulfite as the scrubbing reagent. Davy Powergas developed the process during the 1960's; the first commercial installation was in 1970 on a sulfuric acid plant. The process has since been applied widely on acid plants, Claus plants, and oil-fired boilers and recently has been installed on a coal-fired boiler. It has treated flue gas with SO_2 concentrations from 1000 to 30,000 ppm at 90 percent or better availability.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Flue gases Sodium Sulfite Scrubbing Flue Gas Desulfurization Alternatives	NATO United States	68A 97F

18. DISTRIBUTION STATEMENT RELEASE TO PUBLIC - BEST AVAILABLE COPY	19. SECURITY CLASS *(This Report)* UNCLASSIFIED	21. NO. OF PAGES 67
	20. SECURITY CLASS *(This page)* UNCLASSIFIED	22. PRICE PC A04 MF A01

EPA Form 2220-1 (9-73)

NATO-CCMS STUDY

PHASE I

Flue Gas Desulfurization Pilot Study Phase I

Survey of Major Installations - Sodium Sulfite Scrubbing Flue Gas Desulfurization Process

APPENDIX 95-J

Lead Country - United States

Prepared by

Frank Princiotta, U.S. Environmental Protection Agency, Washington, D.C.
Richard W. Gerstle, PEDCo Environmental, Cincinnati, Ohio
Edmund Schindler, PEDCo Environmental, Cincinnati, Ohio

January 1979

TABLE OF CONTENTS

		Page
SUMMARY		iv
1.0	BACKGROUND AND INFORMATION	1-1
2.0	GENERAL PROCESS DESCRIPTION	2-1
	2.1 Process Chemistry	2-1
	2.2 Description of Equipment Components	2-11
3.0	CURRENT STATUS, ASSESSMENT, AND PLANS	3-1
4.0	RECENT TECHNOLOGICAL DEVELOPMENTS AND SOLUTIONS TO PROBLEM AREAS	4-1
5.0	COST	5-1
6.0	BIBLIOGRAPHY	6-1
APPENDICES		
	A. Questionnaire Data	A-1
	B. Availability Calculation	B-1
	C. Conversion Factors	C-1

LIST OF FIGURES

No.		Page
1	Sodium Solution - SO_2 Reduction Process, Flow Diagram and Material Balance as Applied to a 500 MW Coal-Fired Boiler	2-2
2	Concentrated Active Alkali System: Sulfur Dioxide Vapor Pressure vs. pH at 54°C	4-2
A	Flowsheet of the Wellman-MKK Process (Nishinagoya Station, Chubu Electric Power)	A-17
B	Layout of the FGD Plant at Nishinagoya Station	A-18

LIST OF TABLES

No.		Page
1	Summary of Status of Surveyed Wellman-Lord FGD Systems	3-3
2	Status of Surveyed Installations Wellman-Lord Process	3-4
3	Cost Data For Individual Installations of Wellman-Lord Systems	5-2

SUMMARY

The Wellman-Lord flue gas desulfurization (FGD) process is currently the most recognized regenerable system using soluble sodium sulfite as the scrubbing reagent. Davy Powergas developed the process during the 1960's; the first commercial installation was in 1970 on a sulfuric acid plant. The process has since been applied widely on acid plants, Claus plants, and oil-fired boilers and recently has been installed on a coal-fired boiler. It has treated flue gas with SO_2 concentrations from 1000 to 30,000 ppm at 90 percent or better efficiency and 95 percent or better availability.

The system is best applied to sources in industrialized areas where a market for SO_2, S, H_2SO_4, and Na_2SO_4 exists nearby. No sludge is generated and the only raw material is NaOH or Na_2CO_3 to make up for oxidation of Na_2SO_3 to Na_2SO_4.

In the Wellman-Lord process, SO_2 in the flue gas is contacted with an aqueous solution of sodium sulfite, thereby forming sodium bisulfite. This solution is pumped to an evaporator, where it is heated to release the SO_2. The SO_2 gas then passes through a condenser to remove excess

water and on to a sulfuric acid, elemental sulfur, or liquid SO_2 plant.

As is true with all regenerable sodium solution scrubbing systems, some of the sodium sulfite oxidizes to sodium sulfate in the system. Because sodium sulfate cannot be regenerated easily, it must be removed. Removal of the inactive sulfates is accomplished by pumping a small stream of sodium bisulfite liquor to a refrigerated crystallizer. Here, a sulfate rich crystal is removed from the liquor. The liquor then is returned to a dissolving tank, where the sodium ion removed as sulfate is replenished with sodium hydroxide or sodium carbonate. The sulfates are either sold or neutralized and disposed of. Alternatively, a small stream can be sent to an evaporator-crystallizer, where a sulfate-rich anhydrous crystal is removed. The liquor is returned to the main evaporator. The cleaned flue gas is demisted, reheated, and released to the atmosphere.

Disposal of sodium sulfate may be a problem since it is soluble, and landfilling often is not practical. The salt can be sold where a market is available. Sulfate paper mills are a major user. Davy Powergas is currently investigating methods to convert sodium sulfate back to an active form.

The following are the major advantages and disadvantages of the Wellman-Lord FGD system:

Advantages

1. No scaling takes place in the system.
2. High SO_2 removal efficiencies (95 percent or better) are achievable.
3. Low liquid-to-gas (L/G) ratios are required in the scrubber.
4. The simplicity of the various unit operations enhances the likelihood of reliable performance.
5. Recovered SO_2 can be used to produce high-grade sulfur, sulfuric acid, or liquid SO_2.
6. Considerable operating experience has been obtained on oil-fired boilers, sulfuric acid plants, and Claus plants, and some experience has been obtained on coal-fired boilers.

Disadvantages

1. A small quantity of purge solids (sodium sulfate and other sodium salts) must be sold as a by-product or disposed of. The soluble sodium salts in this purge stream may lead to water pollution problems.
2. The process uses large quantities of low-pressure steam and the high energy demand results in a derating of the power station, often by as much as 6 percent.
3. The evaporator system must be maintained free of solids; therefore, the incoming flue gas must be very low in particulates, the scrubbing solution must be filtered to remove these solids before it is returned to the evaporator, or both.

1.0 BACKGROUND AND INFORMATION

This report is one of a series on the status of flue gas desulfurization processes in the NATO countries and Japan. The United States, the Federal Republic of Germany, and Norway have acted as lead countries in the gathering of information for and preparation of the reports.

The Wellman-Lord process involves the absorption of the SO_2 in flue gas in a solution of sodium sulfite. The sulfite is converted to bisulfite, and the solution is regenerated by thermal decomposition in an evaporator/crystallizer unit. The SO_2 is recovered in a concentrated stream, and it may be used for the production of sulfuric acid, elemental sulfur, or liquid SO_2. This process is applicable to any flue gas containing SO_2. Thirty-one commercial installations are either now in operation or under construction in the United States and Japan.

Davy Powergas developed the Wellman-Lord SO_2 recovery process during the 1960's. Its original purpose was to supply an economical means of recovering sulfur from fertilizer production. In 1965 the original pilot plant work was conducted at Tampa Electric's Gannon Station, using a

potassium-based scrubbing solution. Although another pilot plant (at the Crane Station of Baltimore Gas & Electric) failed because of excessive steam consumption and absorber plugging problems, the experience led to the present sodium sulfite/bisulfite system and rapid commercialization.

The first commercial installation in the United States was on an Olin Chemical sulfuric acid plant in Paulsboro, New Jersey. Most of the early Wellman-Lord systems were smaller units used at sulfuric acid and Claus-type sulfur recovery plants. The process was recently been applied more widely, e.g. on oil-fired boilers, particularly in Japan. Of the 26 systems expected to be operational worldwide by December 1977, 15 are on oil-fired boilers, 7 on Claus plants, 2 on acid plants, and 2 on coal-fired boilers. Treatment capacities of these systems range as high as 739,000 Nm^3/h of flue gas. Sulfur dioxide removal efficiency is typically 90 to 98 percent. Collected SO_2 is recycled at Claus and acid plants and used for the production of additional sulfuric acid, elemental sulfur, or liquid SO_2. Three Wellman-Lord systems are currently under construction in the United States; two are on coal-fired power plants, and one is on a fluid coke-fired steam boiler at a refinery. A particularly noteworthy system was put into operation in June 1971 at the Japan Synthetic Rubber

plant, Chiba, Japan. This system operated at 97 percent availability during its first year and at 100 percent during its second and third years. The system, which requires only one-half man per shift to operate (including production and disposal), has maintained 90 percent SO_2 removal efficiency.

The first Wellman-Lord process FGD unit retrofitted to a coal-fired utility boiler is Unit No. 11 (115 MW) at the Dean H. Mitchell Station of Northern Indiana Public Service Company, Gary, Indiana. On September 14, 1977, the acceptance testing for this unit was successfully completed. All the process performance guarantees were met.

2.0 GENERAL PROCESS DESCRIPTION

2.1 PROCESS CHEMISTRY

The Wellman-Lord process (shown schematically in Figure 1 applied to a 500-MW coal-fired boiler) is based on absorption of SO_2 in an aqueous sodium sulfite/bisulfite solution. The spent absorbant is regenerated in a steam-heated evaporator/crystallizer which releases SO_2. The SO_2 is sent to a recovery plant, where H_2SO_4, sulfur, or liquid SO_2 are produced. A purge stream removes impurities that build up in the absorbant.

2.1.1 Absorption

After the flue gas from the boiler is cleaned of particulate, it is passed through an absorption tower, where contact with aqueous sodium sulfite promotes absorption as follows:

$$SO_2 + Na_2SO_3 + H_2O \rightarrow 2NaHSO_3 \quad (1)$$

Formation of sodium sulfate occurs by oxidation of sodium sulfite.

$$2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4 \quad (2)$$

After absorption, flue gases are demisted, reheated, and vented to the atmosphere. The spent absorber liquor,

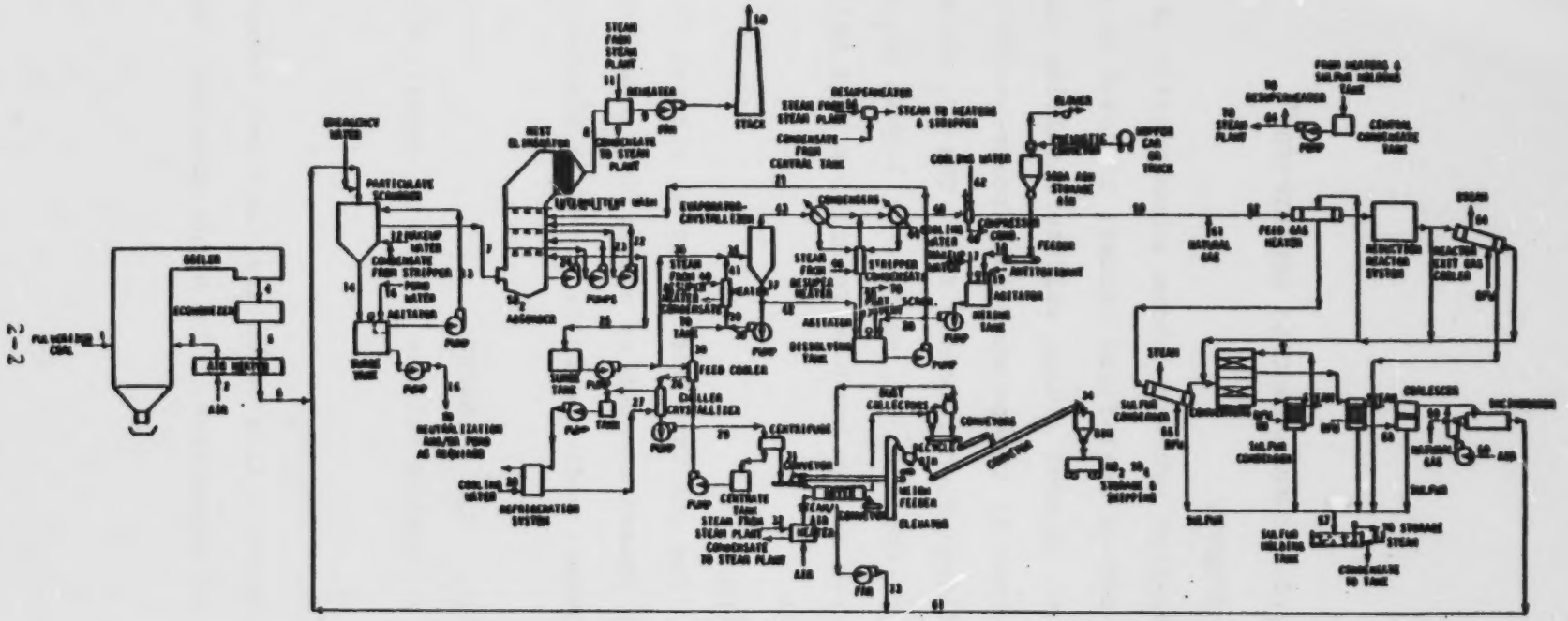

Figure 1. Sodium solution - SO_2 reduction process, flow diagram and material balance as applied to a 500 MW coal-fired boiler.

(See notes at end of figure.)

(continued)

Stream no. / Description	1 Coal to boiler	2 Combustion air to air heater	3 Combustion air to boiler	4 Gas to economizer	5 Gas to air heater	6 Gas to particulate scrubber	7 Gas to SO_2 absorber	8 Gas to reheater	9 Gas to fan	10 Gas to atmosphere
Rate, 1000 kg/h	170	2049	1848	1998	1998	549.8	597.4	595.1	595.1	2380
Nm^3/s		464	419	445	445	123	138	137	137	547
m^3/s, Liquid										
Particulates, kg/h				15,300	15,300	3826	19.5	25.3	25.3	101
Temperature, °C		43.3	279	477	374	154	57.6	57.6	72.8	79.4
Specific gravity										
Viscosity, pascal-seconds										
Undissolved solids, %										
pH										
Sulfur as SO_2, kg/h				11,335	11,335	2834	2645	284	284	1134

Stream no. / Description	11 Steam to gas reheater	12 Makeup water to part. scrbr.	13 Recycle slurry to part. scrbr.	14 Part. scrbr. slurry to surge tank	15 Particulate slurry product	16 Recycle pond water	17 Makeup water to mixing tank	18 Soda ash to mixing tank
Rate, 1000 kg/h	6.8	22.9	1107	1111	25.4	21.6	4.8	1.202
Nm^3/s								
m^3/s, Liquid		0.006	0.2990	0.3000	0.006	0.006	0.0013	
Particulates, kg/h								
Temperature, °C	243			57.6				
Specific gravity			1.03		1.09			
Viscosity, pascal-seconds								
Undissolved solids, %			5		15			
pH				3				
Sulfur as SO_2, kg/h			10,585	10,585	242			

(continued)

Figure 1 (continued).

Stream no.	19	20	21	22	23	24	25	26
Description	Antioxidant to mixing tank	Makeup to dissolving tank	Feed solution to SO_2 absorber	Recycle to third stage	Recycle to second stage	Recycle to first stage end product	Liquor to surge tank	Liquor to chiller-crystallizer
Rate, 1000 kg/h	0.0205	6.03	39.6	0.233	0.276	0.320	42.0	97.1
Nm^3/s								
m^3/s, Liquid		0.0014	0.009	0.051	0.060	0.0695	0.009	0.021
Particulates, kg/h								
Temperature, °C		37.8					134	
Specific gravity		1.20						20
Viscosity, pascal-seconds								
Undissolved solids, %								
pH								
Sulfur as SO_2, kg/h			3543	21,633	26,413	31,410	5905	13,665

Stream no.	27	28	29	30	31	32	33	34	35
Decription	E.C.[a] solution to chiller-crystallizer	Cooling water to refrig. system	Slurry to centrifuge	Centrate to heater	Cake to conveyor	Steam to steam/air heater	Gas from dryer	Purge to bin	Liquor to Evapr. crystr.
Rate, 1000 kg/h	632.8	176	97.1	93.0	4.255	9.25	85.3	1.684	35.4
Nm^3/s							19.2		
m^3/s, Liquid	0.1645	0.049	0.021	0.020					0.008
Particulates, kg/h									
Temperature, °C	-3.9		1.7	37		243	110		
Specific gravity	1.07								
Viscosity, pascal-seconds									
Undissolved solids, %									
pH									
Sulfur as SO_2, kg/h	13,665		13,665	12,907	759			759	4977

(continued)

Figure 1 (continued).

Stream no. / Description	36 Feed to evapr.-crystr.	37 Slurry from evapr.-crystr.	38 Slurry to heater	39 Feed to Heater	40 Steam to Heater	41 Slurry to evapr.-crystr.	42 Slurry to dissolving tank	43 Vapor to primary condenser	44 Water to condenser	45 Steam to stripper
Rate, 1000 kg/h	27,200	27,200	27,200	27,200	52.2	27,200	33.3	49.8	2.53	1.664
Nm^3/s, m^3/s, Liquid	4.33	4.33	4.32	4.33		4.33	0.005	16.7	0.700	
Particulates, kg/h										
Temperature, °C					104					104
Specific gravity										
Viscosity, pascal-seconds										
Undissolved solids, %										
pH										
Sulfur as SO_2, kg/h	$5.8x10^6$	$5.8x10^6$	$5.8x10^6$	$5.8x10^6$		$5.8x10^6$	7086	4345		

Stream no. / Description	46 Condensate to part. scrbr.	47 Condensate to dissolving tank	48 Gas to compressor	49 Condensate	50 Gas to SO_2 reduction	51 Natural gas to SO_2 reduction	52 Mixed gases to feed gas heater	53 Recycle purge to dryer	54 Steam to steam plant
Rate, 1000 kg/h	2.342	43.4	4.85	0.167	9.34	1.500	10.8	25.2	4.3
Nm^3/s			0.59		1.07	0.533	1.60		
m^3/s, Liquid	.0006	.012							
Particulates, kg/h									
Temperature, °C					121	93.3		104	121
Specific gravity									
Viscosity, pascal-seconds									
Undissolved solids, %									
pH									
Sulfur as SO_2, kg/h			4345		8688		8688		

(continued)

Figure 1 (continued).

Stream no. / Description	55 BFW to condensers and gas cooler	56 Steam to desuper-heater	57 Sulfur to storage & shipping	58	59 Natural gas to incinerator	60 Air to incin-erator	61 Gas from incin-erator	62 Cooling water to gas compressor
Rate, 1000 kg/h	6.21	0.102	4.237		0.110	3.998	10.7	34
Mg^3/s					0.03	0.91	2.57	
m^3/s, Liquid	0.007		0.0006					0.009
Particulates, kg/h								
Temperature, °C	110	243				21	677	
Specific gravity								
Viscosity, pascal-seconds								
Undissolved solids, %								
pH								
Sulfur as SO_2, kg/h			8474				214	

Stream no. / Description	63 Steam to sulfur storage	64 Condensate to steam plant
Rate, 1000 kg/h	1.361	103
Mg^3/s		
m^3/s, Liquid		0.028
Particulate, kg/h		
Temperature, °C	148	
Specific gravity		
Viscosity, pascal-seconds		
Undissolved solids, %		
pH		
Sulfur as SO_2, kg/h		

Notes:
1. Calculations based on:
   a. 105% stoichiometric sodium.
   b. 3.5% sulfur in coal (dry).
   c. 12% ash coal (as fired).
   d. 92% of sulfur in coal evolves as SO_2.
   e. 75% of ash in coal evolves as fly ash.
   f. 99.5% removal of particulates to scrubber.
   g. 90% SO_2 removal.
2. Particulates should be added to gas to get total stream rate.
3. Stream numbers 6-9, 11-16, and 21-25 are one of four similar streams.
4. Stream numbers 35-49 and 62 are one of two similar streams.

Figure 1 (continued).

now rich in sodium bisulfite, is divided into two streams -- 90 percent to the evaporator for regeneration and 10 percent to the purge processing system.

2.1.2 Regeneration

The regeneration step occurs in the evaporator. Heat is added to the spent absorber liquor to reverse reaction 1 and to promote crystallization of the less soluble sulfite salts.

$$2\ NaHSO_3 \rightarrow Na_2SO_3\downarrow + H_2O + SO_2\uparrow \qquad (3)$$

The water in the evaporator overhead is partially condensed, stripped of SO_2, and recycled to the process to dissolve the slurry discharge from the evaporator. The SO_2/H_2O stream from the evaporator contains about 5 percent SO_2 by volume and the balance is water.

Makeup Na_2CO_3 or NaOH is added to the dissolving tank, where the stripper and evaporator bottoms are combined to form the new absorber feed. The Na_2CO_3 or NaOH becomes an active absorbant by the following reactions:

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2\uparrow \qquad (4)$$

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O \qquad (5)$$

2.1.3 Product Stream

The vapor stream from the evaporator is first cooled to condense water vapor and to reach 80 to 90 percent SO_2 concentration. This concentrated SO_2 can be used to produce

several end products. Wellman-Lord systems on sulfuric acid plants or Claus plants recycle the gas, whereas units installed on boilers produce sulfuric acid, elemental sulfur, or liquefied SO_2, depending on available facilities. The production of these products is explained below:

- Sulfuric acid - SO_2 is reacted with oxygen in the presence of a vanadium pentoxide catalyst to form SO_3, which is in turn absorbed into 98 percent sulfuric acid. This contact process can make sulfuric acid in any strength desired (usually 93 or 98 percent) or oleum. A single contact plant can be used.

  The tail gas is recycled to the Wellman-Lord absorber.

- Liquid SO_2 - The SO_2 gas is dried with silica gel or sulfuric acid, then compressed and condensed. The resulting liquid SO_2 is collected and stored in a pressurized tank.

- Elemental sulfur - The process for recovery of elemental sulfur with a minimum assay of 99.5 percent is carried out in two steps. In a reduction section, a portion of the SO_2 reacts with methane and produces a mixture of elemental sulfur, hydrogen sulfide, carbon dioxide, water vapor, and some unreacted SO_2. (Propane or other similar reducing gases can be used as the reductant.)

$$2CH_4 + 3SO_2 \rightarrow 2CO_2 + 2H_2O + S + 2H_2S \qquad (6)$$

  In the Claus section of the process, the hydrogen sulfide and remaining SO_2 from the reduction section react to produce additional elemental sulfur and water vapor.

$$2H_2S + SO_2 \rightarrow 2H_2O + 3S\downarrow \qquad (7)$$

The tail gas is incinerated and recycled to the absorber.

The theoretical quantities of each end product generated can be calculated as follows:

$$H_2SO_4: \frac{98\ g/mol\ H_2SO_4}{64\ g/mol\ SO_2} \cdot \frac{0.11\ kg\ Na_2SO_4\ purged}{kg\ SO_2\ removed} \times \frac{98\ g/mol\ H_2SO_4}{142\ g/mol\ Na_2SO_4} = \frac{1.46\ kg\ H_2SO_4\ end\ product}{kg\ SO_2\ removed}$$

$$SO_2: \frac{1\ kg\ SO_2}{kg\ SO_2\ removed} \cdot \frac{0.11\ kg\ Na_2SO_4\ purged}{kg\ SO_2\ removed} \times \frac{64\ g/mol\ SO_2}{142\ g/mol\ H_2SO_4} = \frac{0.95\ kg\ SO_2\ end\ product}{kg\ SO_2\ removed}$$

$$S: \frac{32\ g/mol\ S}{64\ g/mol\ SO_2} \cdot \frac{0.11\ kg\ Na_2SO_4\ purged}{kg\ SO_2\ removed} \times \frac{32\ g/mol\ S}{142\ g/mol\ Na_2SO_4} = \frac{0.47\ kg\ S\ end\ product}{kg\ SO_2\ removed}$$

Reproduced from best available copy.

2.1.4 Purge Stream

Nonregenerable sodium sulfate is also formed in the absorber by oxidation of sulfite or by absorption of sulfur trioxide:

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4 \quad (8)$$

$$2\ Na_2SO_3 + SO_3 + H_2O \rightarrow Na_2SO_4 + 2\ NaHSO_3 \quad (9)$$

Continuous purge limits the amount of salt cake generated to 5 percent of the sulfur removed. If the salt cake is assumed to be 100 percent Na_2SO_4, then 0.11 kg Na_2SO_4/kg SO_2 is generated. Thiosulfate and dithionate compounds also may be present in small quantities. Another source of sodium sulfate and another nonregenerable product, sodium thiosulfate ($Na_2S_2O_3$), occurs in the regeneration section as follows:

$$6\ NaHSO_3 \rightarrow 2\ Na_2SO_4 + Na_2S_2O_3 + 2SO_2 + 3H_2O \quad (10)$$

Sodium sulfate and sodium thiosulfate from the bisulfite-rich absorber stream, are removed in a crystallizer. The solids are dried either for marketing or disposal, or they are acidified, neutralized, and then discharged. Soda ash (Na_2CO_3) or caustic soda (NaOH) is added to the dissolving tank to make up for sodium lost in the purge stream. If a 5 percent loss is assumed, soda ash makeup would be 0.083 kg/kg SO_2 removed.

$$0.05 \times \frac{106 \text{ g/mol } Na_2CO_3}{64 \text{ g/mol } SO_2} = 0.083 \text{ kg } Na_2CO_3 \text{ added per kg } SO_2 \text{ removed.}$$

One of the principal process problems is the formation of sodium sulfate from both the absorption of SO_3 and the oxidation of sodium sulfite by oxygen. It is difficult to reverse this action to rejuvenate the inactive sodium; therefore, inactive sodium salts must be purged from the system and their sodium ions replaced by active sodiums ions in the form of soda ash or caustic soda. Thus, this purge involves both a chemical makeup requirement and a potential waste disposal problem. The purge stream can remove 5 to 10 percent of the total sulfur, and treatment is required to prevent water pollution. The amount of the purge stream has increased proportionately with increasing sizes of new Wellman-Lord installations. Attempts have been made to prevent solution oxidation by the addition of different

antioxidants, but this approach has been abandoned because the cost of inhibitors is high and some purge stream would still be required. Selective removal of the oxidized part of the solution by crystallization (by chilling) has been more successful and has resulted in a five- to sixfold decrease in the purge stream and chemical makeup requirements. Concentrated sodium sulfate crystals in the purge stream can be dried for sale or disposal, or they can be further treated and discharged into a watercourse. This crystallization process is used at many Wellman-Lord installations.

2.2 DESCRIPTION OF EQUIPMENT COMPONENTS

2.2.1 Particulate Removal

Before the sulfur oxides are removed, flue gas from boilers is passed through an electrostatic precipitator (ESP) and then, in most cases, through a venturi scrubber. The ESP is designed to remove 97 to 99 percent of the particulates. The venturi scrubber removes chlorides and additional particulates and cools and humidifies the flue gas before the absorption process. This not only prevents particulate emissions, but also reduces contamination of the absorbing liquor and of any salable sodium sulfate by-product.

2.2.2 SO_2 Absorber

The SO_2 absorber is a simple, gas-liquid contact device. The shell can be constructed of tile-lined concrete, plastic-lined mild steel, or stainless steel. The lean sodium sulfite scrubbing solution enters the top of the column, circulates countercurrently over sieve, valve trays, or packing, and contacts the ascending gas. Depending on the inlet SO_2 concentration and the design removal efficiency, a tower can contain three to five stages. Koch Multiventuri Flexitrays or Glitsch Ballast Trays of 316L stainless steel are used. Each tray adds 750 to 870 Pa pressure drop to the system, and the absorber module can be designed to accommodate more trays. Each stage has one pump; no spares are required because the other stages can handle an increased load with very little loss of efficiency until a worn pump is replaced. Because of the extremely low L/G ratio required for absorption, the absorbing solution is recirculated on each set of trays. Recirculation of this absorbing solution also provides the necessary liquid to load the valve trays hydraulically. As the solution chemically absorbs SO_2, forming sodium bisulfite, it becomes more soluble, allowing for nonfouling, scale-free operation. The scrubber solution, rich in bisulfite, is discharged into a surge tank.

The quantity of flue gas to be treated is the most important design parameter in the sizing of the absorber system. Because of the successful operating history of the Wellman-Lord system (better than 96 percent on-stream time for the absorption area), Davy Powergas does not recommend the use of spare absorbers. A modular concept is recommended, and the present maximum size blower, prescrubber, and absorber module has a gas handling capacity equivalent to 150 to 185 MW for coal-fired power plants.

Concentrations of SO_2 in the feed gas can range widely. The higher absorption rate of stronger SO_2 streams allows higher removal efficiency with fewer stages. For example, flue gas with an SO_2 concentration of 1200 ppm requiring 90 percent removal will need four to five stages, whereas a gas concentration of 3000 ppm requiring 95 percent removal may only need three stages. Each mass transfer stage or tray adds 1 to 2 m to the absorber height. The volume of the scrubbing system increases with decreasing SO_2 concentration because the rate of absorbing solution circulation is higher.

The water content of the flue gas also directly affects the number of stages required for SO_2 removal. The SO_2 absorption section operates at a temperature very near the adiabatic saturation temperature of the flue gas. As the water content of the flue gas increases for a given dry bulb

temperature, the increased saturation temperature raises the absorber operating temperature, affecting SO_2 equilibrium. A 2 percent change in volume of water content in the flue gas may require one additional mass transfer stage.

2.2.3 Demister

The demister downstream from the absorber is necessary to remove entrained solution droplets, thereby decreasing downstream equipment corrosion and decreasing the energy requirement in the reheater and any chemical loss. Most of the droplets are large enough to be removed by a simple change in gas direction. Chevron demisters are sufficient, and they can be placed in series for additional capacity. Demisters may tend to foul, however, so spray washes are installed to reduce this tendency. Demisters can be constructed of stainless steel, fiberglass-reinforced polyester, or other plastic material, and can be installed in the horizontal or vertical position. The vertical position (see Figure 1) has been found to be more effective in this process, as the horizontal demister allows captured water to drip back in the gas stream and become re-entrained.

2.2.4 Stack Gas Reheater

Reheating of stack gas is generally necessary to maintain plume dispersion. Secondary advantages of reheating are that it protects downstream equipment such as ducts

and stack and prevents acid mist formation. Reheating temperatures generally range from 28° to 45°C and can be accomplished by 1) installing a gas or low-sulfur oil burner that exhausts directly into the stack, 2) installing steam coils to heat flue gas, 3) using steam coils to heat air and injecting it into the stack, or 4) by passing untreated hot flue gas. An in-line heat exchanger costs the least initially, but maintenance costs are high because it is subject to corrosion.

2.2.5 Evaporation Crystallizer

The bisulfite-rich solution from the surge tank flows at a steady rate into the evaporator crystallizer constructed of 316L stainless steel. The solution is thermally decomposed in the evaporator, using process steam, to form sodium sulfite, water, and SO_2. The water and SO_2 are driven off overhead, and the sodium salts remaining in the circulating liquid precipitate out to form a dense crystalline slurry composed primarily of sodium sulfite. The evaporator can be heated by low-pressure exhaust steam. The mixture of SO_2 and water vapor is concentrated through a series of partial condensers, the first of which may be the heat exchanger for a second-stage evaporator. In larger installations, the regeneration system is operated as a double-effect evaporator, reducing steam consumption by 40 to 45 percent. The

condensers may be air- or water-cooled. The condensed water, which contains varying small amounts of dissolved SO_2, is sent to a stripper and returned to a dissolving tank. The saturated gas, normally about 85 percent SO_2 by volume, goes to an end product plant.

2.2.6 Stripper

The stripper separates SO_2 from condensed H_2O, thereby increasing the amount of SO_2 supplied to the end product plant. The stripper can be a tray-type or packed column, directly heated by steam at the bottom.

2.2.7 Dissolving Tank

Slurry from the evaporator is combined with the stripper condensate to form the sodium sulfite solution feed for the absorber. Sodium hydroxide or soda ash is added to compensate for the sodium ions lost in the sulfate purge stream; then the solution flows to a surge tank for recycling back to the absorber.

2.2.8 Surge Tanks

Surge tanks on the absorption and regeneration sections stabilize liquid flow to the regeneration plant by minimizing flow fluctuations in the system and allowing the absorption and regeneration sections to be located some distance from each other. They also permit the regeneration system to be shut down completely for up to 3 days without

interrupting SO_2 removal in the absorber, or they allow the boiler to be shut down for up to 3 days without interrupting a minimal flow of SO_2 to the end product plant. The boilers in one installation are down over the weekend, but the H_2SO_4 plant runs continuously.

2.2.9 Chiller Crystallizer

To prevent the buildup of sodium sulfate and other inert materials, some of the liquid leaving the absorber must be purged from the system. Because it has a high bisulfate content combined with a low sulfite content and consequently is amenable to sulfate separation, a small sidestream is taken from the absorber product stream. Most of the product solution is returned to the evaporator for regeneration. This purge stream can represent 5 to 10 percent of total sulfur removal. It is cooled in a chiller/crystallizer to 0°C; the sulfate precipitates first because it is the least soluble of the sodium compounds in the solution. The sulfate-rich solid phase from this controlled crystallization and any other solids in the solution are separated from the liquid in a centrifuge. The liquid is returned to the dissolving tank, where the bisulfite combines with the Na_2CO_3 or NaOH and forms sodium sulfite. The solid phase is 70 to 80 percent Na_2SO_4.

2.3 OPERATING PARAMETERS

2.3.1 Chemical Parameters

The pH of Wellman-Lord absorption systems is maintained between 6 and 7. A pH of 6.5 is the normal control point. A higher pH increases carbon dioxide absorption, whereas a lower pH increases SO_2 absorption. A system with a pH of 6.5 is highly buffered and can adapt rapidly to changes in flue gas inlet conditions. A pH of approximately 5.5 is normally found at the absorber discharge.

2.3.2 Physical Parameters

The primary scrubber in the system can be a particulate scrubber or just a gas cooler and humidifier. Flue gas temperature is normally lowered to 50° to 55°C. Liquid-to-gas ratios for representative systems are 0.5 to 1 liter/Nm^3, flue gas velocities across the throat are about 2 m/s, and the pressure drop across the throat is from 1 to 3 kPa.

The SO_2 absorber has an overall L/G ratio of 0.067 liter/Nm^3, whereas the L/G for each tray is about 0.5 liter/Nm^3. Internal scrubber flue gas velocities are about 3 m/s. Pressure drops in the absorber vary from 1 to 4 kPa, depending on the number of trays.

Flue gas velocities in a vertical demister average 6 m/s. The optimum flue gas velocity in horizontal demisters is about 3 m/s. Increasing the velocity increases the water

removal rate, but it also increases re-entrainment of water before it can leave the demister. Pressure drop across the demister ranges from 50 to 150 Pa.

Flue gas reheating temperatures at representative boilers range from 28° to 45°C and energy expenditure approximates 1100 J/kg of flue gas per °C. Pressure drop across the reheater is 250 to 700 Pa.

The purge rate of dry Na_2SO_4 cake is about 62 kg/h at a 100-MW station for each 1 percent of O_2 in the flue gas. In a system where the Na_2SO_4 is purged as a neutralized wastewater, about 1 kg of water is needed as makeup for each mole of SO_2 removed.

3.0 CURRENT STATUS, ASSESSMENT, AND PLANS

The Wellman-Lord process has proved itself on a variety of SO_2 emission sources. High removal efficiency and high availability have been shown on oil-fired boilers, Claus plants, H_2SO_4 plants, and most recently, on a coal-fired boiler.

Its best application is in the treatment of gases containing 0.15 to 3 percent SO_2 by volume. From 90 to 95 percent SO_2 removal capability can be achieved on both high- and low-sulfur coal applications. Because of internal recirculation loops, about 50 ppm is the minimum outlet concentration economically attainable. This value is based on the partial pressure of SO_2 over the scrubber liquor. The best location for the Wellman-Lord process is near an industrialized area where markets for sulfur by-products (S, SO_2, H_2SO_4) exist. Since land is not required for disposal, land costs are not an important consideration. An additional advantage can be gained if a market for Na_2SO_4 exists nearby. This product may be landfilled, neutralized and dumped at sea, regenerated, or sold to pulp and paper mills.

Plans are now under way for the installation of several new Wellman-Lord systems. Three of these are in the United States, and all are on coal-fired boilers. The size of the boilers ranges from 100 to 1100 MW. The first application of a Wellman-Lord system on a full-scale, coal-fired utility boiler was started just recently on Mitchell Unit 11 of the Northern Indiana Public Service Corporation (NIPSCO). The success of this jointly funded EPA-NIPSCO system has shown the viability of applying this system to utility coal-fired boilers.

No new FGD systems are being planned or built in Japan because oil desulfurization is less costly than FGD and ambient SO_2 standards are being met.

Table 1 summarizes the status of surveyed Wellman-Lord systems. The collection efficiency and availability ranges refer only to the surveyed systems. The rest of the information refers to all Wellman-Lord systems projected through December 1977. Table 2 presents important characteristics of each surveyed facility, and the questionnaires in the Appendix provide more detailed information.

Table 1. SUMMARY OF STATUS OF SURVEYED WELLMAN-LORD FGD SYSTEMS

All Wellman-Lord Systems	
Total number of systems in operation by December 1977	26
Total number of systems built	27[a]
Total number of systems surveyed	2
Year first placed in commercial service	1970
Number of utility systems using process	4
Total MW of these facilities	810
Number of industrial combustion systems	13
Surveyed Wellman-Lord Systems	
Range of average SO_2 removal efficiencies, %	90-92
Range of fly ash removal efficiency, %	88.-99.5
Range of average availability factor,[b] %	97-99.5

[a] Includes industrial process applications.

[b] See Appendix B.

Table 2. STATUS OF SURVEYED INSTALLATIONS WELLMAN-LORD PROCESS

Name of plant	Location	Size, MW	Fuel type	Characteristics		Flue gas treated,		Year FGD systems started	SO_2 removal efficiency, %	Disposition of end products	Energy required, % total input	Avg. availability, %
				S, %	Heat value, 10^7 J/kg	Nm^3/s	% of total					
Chuba Electric Power Co. Nishinagoya	Asukumura, Kaibegun, Aichi Prefecture	220	Heavy Oil C	2.5	4.37	172 @ 105°C	100	1973	92	H_2SO_4 sold Na_2SO_4 dumped into sea	2.7[a]	99.5
Northern Indiana Public Service Co. Mitchell	Gary, Indiana USA	115	Coal	3.5	2.563	141 @ 142°C	100	1976	90	Sulfur sold Na_2SO_4	9.3[b]	97[c]

[a] In addition, 6.1% reheat required. Reheat in Japan depends on local conditions, not the process.

[b] In addition, 1.3% required for sulfur recovery plant and 2.6% required for reheat.

[c] 62-day period.

4.0 RECENT TECHNOLOGICAL DEVELOPMENTS AND SOLUTIONS TO PROBLEM AREAS

Sodium-based scrubbing processes are not subject to the inherent plugging and scaling problems and limited removal capabilities associated with direct lime- and limestone-based scrubbing processes because the high solubility of sodium alkalis and their sulfur-containing reaction products tend to minimize these problems.

4.1 pH CONTROL

The scrubbing solution pH range provides the key to optimal SO_2-removal efficiency. A control value of 6.5 permits maximum SO_2 removal efficiency without chemical-related problems. If the scrubbing solution pH value exceeds 7, carbon dioxide absorption becomes significant. A scrubbing solution pH level below 5 must also be avoided; it causes the vapor pressure of SO_2 to increase on sulfite-bisulfite systems and can lead to equilibrium-limited scrubbing conditions where outlet concentrations at or below 200 ppm are desired. Graphical evidence for this chemical phenomenon is provided in Figure 2, in which the sulfur dioxide vapor pressure is plotted as a function of pH for a

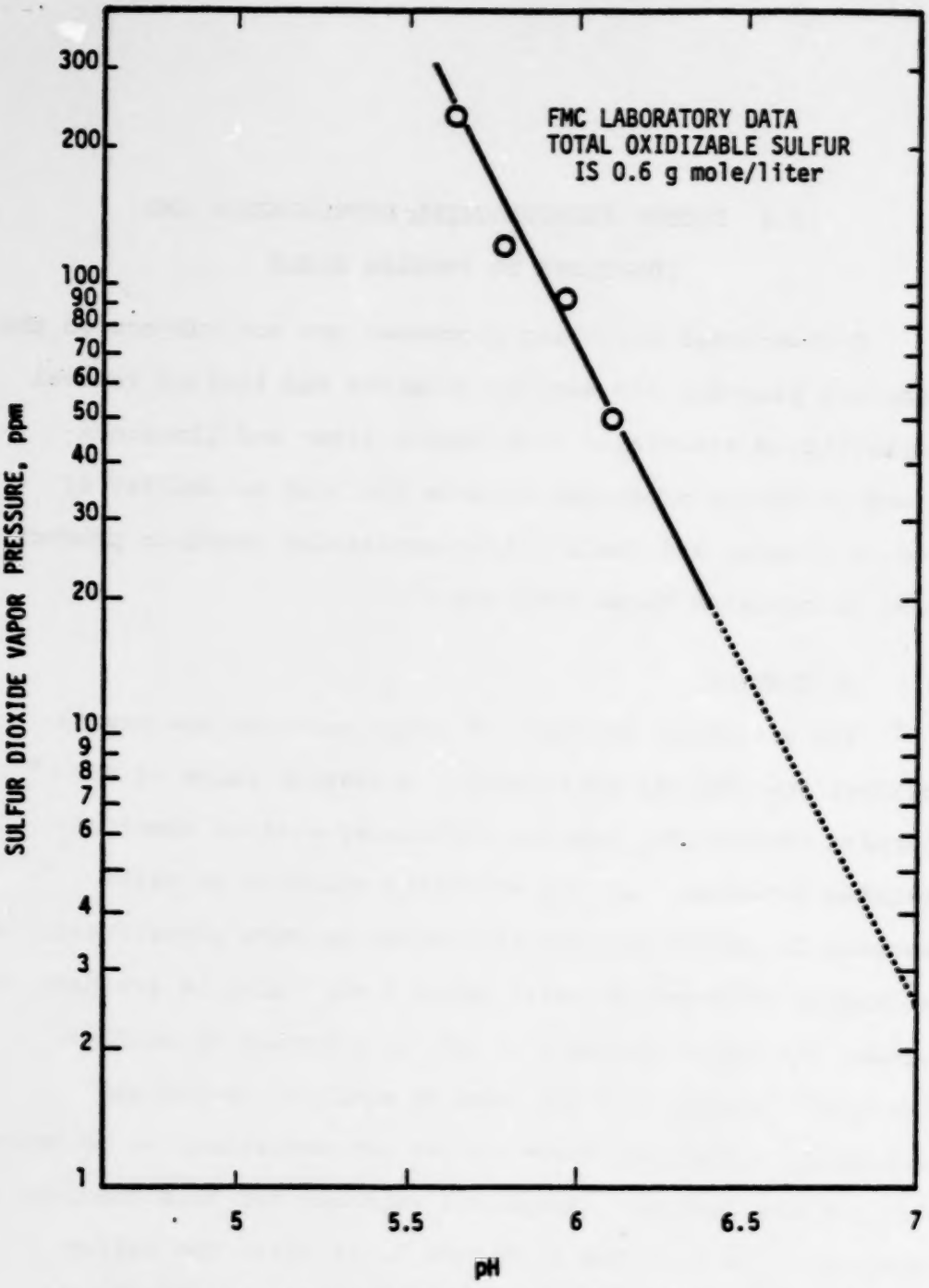

Figure 2. Concentrated active alkali system: sulfur dioxide vapor pressure vs. pH at 54°C.

(From: the FMC Concentrated Double-Alkali Process, by Legatski L.K., K.E. Johnson, and L.Y. Lee, FMC Corp. Glen Ellyn, Illinois)

solution temperature of 54°C, a typical saturation temperature for a boiler flue gas stream. For an active alkali system, the vapor pressure of SO_2 as a function of pH is significant. This is evident from the line in the figure, which was calculated from Henry's constant for SO_2 and water and from the first and second ionization constants for sulfurous acid (H_2SO_3) for a solution with a total oxidizable sulfur (TOS) level of 0.03 g mol/liter and an ionic strength of 0.3 g mol of Na^+/liter. The SO_2 vapor pressure at a pH of 6 is approximately 100 ppm and increases rapidly as the pH drops.

For these reasons, Wellman-Lord systems tend to operate in the pH range of 6 to 7, preferably at a set point of 6.5, where the sulfite-bisulfite system is highly buffered. The highly buffered system is preferred because the scrubbing solution is able to adapt to rapid changes in flue gas inlet conditions. In contrast, scrubbing systems operating at a pH of 6 or less can be very sensitive to and adversely affected by rapid changes in flue gas inlet concentrations.

Based on the relatively low SO_2 vapor pressure in solution at the optimum pH control value of 6.5, SO_2 concentration levels can be maintained at about 50 ppm in the exit gas stream.

As has been noted, all inactive sodium salts must be purged from the scrubber solution and their sodium ions

replaced by active sodium ions from soda ash or caustic soda. Methods are being researched for eliminating the purge stream by reactivating the sodium ion. Several processes have been investigated in both laboratory and full-scale applications, but few details have been disclosed. In one process, the purge stream would be incinerated to liberate the SO_2 and recover the sodium salts, a concept similar to the recovery boiler used in the pulp and paper industry. Two processes are under development to reduce the sodium sulfate chemically. One uses natural gas and the other uses coal as the reductant. Pilot plant testing has been successful. A method employing high-temperature separation of the sulfate also has been tested successfully on a bench scale.

The recovery of elemental sulfur could be limited by natural gas supplies. A 500-MW plant would require about 1360 Nm^3/h of natural gas, depending upon the sulfur content of the coal and the efficiency of the sulfur recovery plant. High-sulfur residual oils, coal gasification, and alternative hydrocarbon fuels such as propane are being investigated as alternate fuels.

The presence of fly ash and particulate matter in the flue gas from a coal-fired boiler presents a potential operating difficulty for a Wellman-Lord installation.

Particulate matter must be kept away from the scrubbing solution because large quantities of particulate will affect the system's operating efficiency.

4.2 CHLORIDES

Chlorides in the scrubbing solution occur primarily from the combustion of chloride bearing fuel and to lesser extent from chloride in the raw water and limestone. Depending on the amount of these chlorides, problems in the scrubber system may occur due to corrosion, decreased SO_2 absorption, and waste water discharge. Chloride concentrations in the scrubber liquor also increase with the amount of liquor recycled. These problems can be reduced by 1) using a prescrubber and keeping that liquid stream separate from the SO_2 absorber streams, 2) designing for chloride corrosion resistance, and 3) increasing the system purge rate. Depending on wastewater discharge regulations, the purge stream may require further treatment.

5.0 COST

A comparison of economics between systems in various countries is complex because economic factors such as labor costs, raw material costs, utility costs, taxes, and interest rates differ, and the differences change from year to year. The inflation rate also varies by country, and the relative value of money changes. A further complication arises because FGD systems are built differently. Japanese systems, for example, have pH controllers and convert all the sludge to gypsum; spares are included in some, but not all systems. This makes it hard to compare two plants even within the same country.

Available cost data are presented in Table 3. The costs have been escalated to reflect a January 1976 initiation of plant construction; the Chemical Engineering plant cost index can be used to compare the 1976 cost of plant construction with the cost during the 1957-59 period and to determine the cost of any point in between. Actual costs are given in the questionnaires in the Appendix.

Chuba Electric provided totals for capital costs, fixed costs, and variable costs in their questionnaire. Itemized

Table 3. COST DATA FOR INDIVIDUAL INSTALLATIONS OF WELLMAN-LORD SYSTEMS

Company name Plant name Location	Chuba Electric Nishinaguya Asukamura, Kaibegun, Aichi Prefecture, Japan	Northern Indiana PS D. H. Mitchell Gary, Indiana, USA
1. Reported capital costs, $	6,000,000	14,824,000[a]
Year	1972	1976
Cost inflated to 1976	8,560,000	14,824,000
$/kW	39	129
2. Reported annual costs, $		
Total variable costs	4,700,000[b]	2,513,000
Fixed charge factor, %	25.	25.98
Total fixed costs	1,500,000[b]	3,854,000[c]
3. Boiler operating hours	7000	
Capacity factor	0.744	
Generated MWh	1,438,837	
4. Reported cost per kWh (mills/kWh)	4.31	8.1
Fixed cost	1.04	
Variable costs	3.26	
Total cost (mills/kWh)[d]		
4000 h	4.75	12.8
8000 h	3.90	8.1

[a] Yen is converted to dollars in year of construction initiation and then converted to January 1976 dollars using the Chemical Engineering Plant Index (1957-59 = 100, Jan. 1976 = 187.1), 1 dollar = 300 Yen.

[b] Estimated.

[c] Fixed cost are normally based on 7 years' depreciation in Japan. This estimate is based on 20 years.

[d] Based on capacity factor of 100%.

costs were not available for maintenance, labor, utilities, raw materials, and product credits. NIPSCO provided maintenance and product credit cost.

All costs reported in the questionnaires were assumed to apply to the date of plant start-up unless otherwise indicated. Plant start-up date and length of construction were generally reported. Annual costs were itemized, but capital costs were not.

Because available cost data were very limited, missing annual costs were estimated.* Reported costs such as electricity and limestone costs were used where available in the estimating procedure, otherwise the following costs were assumed:

Electricity	25 mills/kWh
Limestone	$16.5/Mg
Labor	$10/h

The life of an FGD was assumed to be 20 years; interim replacement, taxes, insurance, and capital charges totaled 0.1365 of the total 1976 capital cost.

---

* Ponder, T.C., L.V. Yerino, V. Katari, Y. Shah, and T.W. Devitt. Simplified Procedures for Estimating Flue Gas Desulfurization Systems Costs. Industrial Environmental Research Laboratory. Research Triangle Park, North Carolina. EPA-600/2-76-150. U.S. Environmental Protection Agency, Washington, D.C., June 1976.

6.0 BIBLIOGRAPHY

1. Ando, J., and G.A. Isaacs. SO_2 Abatement for Stationary Sources in Japan. Industrial Environmental Research Laboratory, Research Triangle Park, North Carolina. Contract No. 68-02-1321, Task 6, Report No. 21ACX-130, Program Element No. 1AB 013. U.S. Environmental Protection Agency, Washington, D.C., January 1976.

2. PEDCo Environmental, Inc., Cincinnati, Ohio. January-February-March report on the Status of Flue Gas Desulfurization Systems in the United States. Industrial Environmental Research Laboratory/Research Triangle Park and the Division of Stationary Source Enforcement, U.S. Environmental Protection Agency, April 20, 1977.

3. Mann, E.L., R.C. Christman, S.F. LaKatos, and A.W. Michener, Jr. Status of Demonstration of the Wellman-Lord/Allied FGD System at Nipsco's D.H. Mitchell Generating Station. U.S. Environmental Protection Agency. (Presented at Symposium on Flue Gas Desulfurization/Regenerable Processes Session. New Orleans, Louisiana. March 8-11, 1976.) 4 p.

4. The McIlvaine Company. The McIlvaine Scrubber Manual, Vol. 3. Northbrook, Illinois, 1974-1977.

5. Slack, A,V., and G.A. Hollinden. Sulfur Dioxide Removal from Waste Gases. Noyes Data Corporation, Park Ridge, New Jersey, 1975. 294 p.

6. PEDCo Environmental, Inc., Cincinnati, Ohio. SO_2 Abatement for Stationary Sources in Japan. Cincinnati, Ohio, June 1976.

7. PEDCo Environmental, Inc., Cincinnati, Ohio. Survey Report on SO_2 Control Systems for Non-Utility Combustion and Process Sources. Contract No. 68-02-2603, Task No. 4. U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, May 1977.

8. Zada, F.K. Status of Flue Gas Desulfurization Processes. PEDCo Environmental, Inc., Cincinnati, Ohio. Aluminum Company of America, Pittsburgh, Pennsylvania, January 13, 1975. 128 p.

9. Ponder, T.C., L.V. Yerino, V. Katari, Y. Shah, and T.W. Devitt. Simplified Procedures for Estimating Flue Gas Desulfurization Systems Costs. Industrial Environmental Research Laboratory. Research Triangle Park, North Carolina. EPA-600/2-76-150. U.S. Environmental Protection Agency, Washington, D.C., June 1976.

10. Standard For Metric Practice. American Society for Testing and Materials, Philadelphia, Pennsylvania, February 1976.

APPENDIX A

APPENDIX A. QUESTIONNAIRE DATA ON THE WELLMAN-LORD FLUE GAS DESULFURIZATION PROCESS

		MITSUBISHI KAKOKI KAISHA (MKK) Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
1.	GENERAL DESCRIPTION		
1.1	Company	Chubu Electric Power Co.	Northern Indiana Public Service Co.
1.2	Plant name	Nishinagoya	D. H. Mitchell Plant
1.3	Preparer's name and mailing address	Jumpei Ando 1-13 Kasuga Bunkyo-Ku, Tokyo	Michael D. Giordano 11499 Chester Road Cincinnati, Ohio 45246
1.4	Principal plant products	Electrical power	Electrical power
	Brief description of area surrounding plant site and factors for choice of FGD system	Located 10 km north of Nagoya and is on reclaimed land at Ise Bay. There are no houses within 5 km of plant. Boiler No. 1 is designed for high sulfur oil while the others are designed for low sulfur.	N.A.
1.5	Plant location	Asahamura, Kaibegun - Aichi Prefecture	Gary, Indiana
1.6	STATION CAPACITY		
1.6.1	Total boiler generating capacity MW/number of boilers	2190/6	115/N.A.
1.6.2	Capacity served by FGD system MW	220	115
	amount of flue gas treated, Nm^3/h	N.A.	358,000 (normal flow at 92 MW)
	% of boiler flue gas treated by FGD system	100	100
	Number of boilers served	1	1

N.A. - Not available.

Nm^3/h - m^3/h (0°C and 101.325 kPa)

(continued)

APPENDIX A (continued)

		MEK Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
1.7	TYPE OF FGD SYSTEM	Wellman Lord-MEK Process	Wellman Lord
1.8	Number of scrubber trains in parallel	1	1
1.9	FGD system installed on new or existing boiler	Existing	Existing
1.10	STATUS OF FGD SYSTEM Operational or under construction, date started	Operational, September 1973	Operational, July 1976
2.	BOILER DATA		
2.1	Boiler identification number	No. 1	No. 11
2.2	Maximum continuous heat input, MW	569	N.A.
2.3	Max. cont. net generating capacity, MW	200	115
2.4	Flue gas flow rate, Nm^3/h temperature at induced draft fan, °C	620,000 105	358,000 142
2.5	Boiler manufacturer		
2.5.1	Fuel type	Heavy oil C	Coal
2.5.2	Firing configuration	Opposite firing	N.A.
2.6	Year placed in service	1970	N.A.
2.7	Type service	Middle	Peaking
2.8	Stack height, m	150	51.2
2.8.1	Flue gas exit velocity, m/s	13.5	18.3
2.9	Average boiler operation, h/yr	7000	N.A.
2.10	Average boiler capacity factor	0.744	N.A.

(continued)

APPENDIX A (continued)

		MHI Hishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
2.11	COAL COMPOSITION	N.A.	
2.11.1	Lower heating value kJ/kg		25,586
2.11.2	Sulfur content, %		3.2-3.5
2.11.3	Ash content, %		10.6
2.11.4	Chloride content, %		0.05
2.11.5	Moisture, %		11.0
2.12	OIL COMPOSITION		
2.12.1	Grade/heating value, kJ/kg	Heavy oil C/43,750	N.A.
2.12.2	Sulfur content, %	2.5	
2.12.3	Ash content, %	N.A.	
2.12.4	Chloride content, %	N.A.	
2.13	PARTICULATE REMOVAL EQUIPMENT NOT PART OF FGD SYSTEM		
2.13.1	Type	Electrostatic Precipitator	Electrostatic Precipitator
2.13.2	Manufacturer	Hitachi Ltd.	N.A.
2.13.3	Efficiency, % removal	80	N.A.
2.13.4	Emission rate, mg/Nm^3	50	700

(continued)

APPENDIX A (continued)

		MKK Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
3.	EMISSION REGULATIONS		
3.1	ATMOSPHERE EMISSIONS		
3.1.1	SO_2	23.6 Mg/day (from entire plant)	0.516 ng/kJ
3.1.2	Particulate	50 mg/Nm^3 (No. 1) 30 mg/h (Nos. 2-6)	0.043 ng/kJ
3.1.3	NO_x	28.1 Mg/day (from entire plant)	N.A.
3.1.4	Other	N.A.	N.A.
3.1.5	Legal basis	Agreement with Asukamura (town)	Federal standard
3.2	RESTRICTIONS APPLICABLE TO LIQUID OR SOLID WASTES, LEGAL BASIS	pH 7.0-8.3 COD 10 ppm SS 15 ppm wastewater Oil 1 ppm Agreement with Asukamura	
4.	FGD SYSTEM INFORMATION		
4.1	Process name	Wellman Lord MKK Process	Wellman Lord
4.1.1	Supplier or design company	Mitsubishi Kakoki Kaisha (MKK)	Davy Powergas, Inc.
4.1.2	WHY WAS THIS SYSTEM CHOSEN?	Reliability of operation and usefulness of the by-product sulfuric acid.	N.A.
4.1.3	PROCESS DESCRIPTION	Appendix Figure A and B	SO_2 absorbed by Na_2SO_3, which is regenerated and yields a concentrated stream of SO_2 to be reduced to elemental sulfur.

(continued)

APPENDIX A (continued)

		MHI Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
4.2	SO_2 SCRUBBING AGENT		
4.2.1	Name	NaOH (Na_2SO_3)	Sodium sulfite
4.2.2	Chemical composition	NaOH - H_2O	Na_2SO_3
	Components, % by weight	NaOH, 45 - H_2O, 55	Na_2SO_3, 17.28 (in feed soln.)
4.2.3	Stoichiometric requirement	N.A.	N.A.
4.2.4	Addition rate, ℓ/h	400	N.A.
4.2.5	Particle size, mm	N.A.	N.A.
4.3	REGENERATION AGENT		
4.3.1	Name		Sodium bisulfite
4.3.2	Composition, %		Na_2HSO_3, 21.0 (in prod. soln.)
4.3.3	Stoichiometric requirement		N.A.
4.3.4	Addition rate		N.A.
4.3.5	Particle size		N.A.
4.4	SALABLE PRODUCT		
4.4.1	Name	Sulfuric acid	Sulfur
4.4.2	Chemical composition Components, % by weight	H_2SO_4, 98 H_2O, 2	S, 99.5 (min)
4.4.3	Amount generated/year, ton/yr	14,500 (approx.)	6298
4.4.4	Sales potential	All has been sold.	Total sale.

(continued)

APPENDIX A (continued)

		MKK Mishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
4.4.5	BY-PRODUCT RECOVERY PROCESS		N.A.
	Name	Sulfuric acid recovery plant	
	Vendor	MKK	
	Process description	Contact process	
4.4a	Intermediate product composition	SO_2 gas	
4.5	BY-PRODUCT No. 2	N.A.	Purge salts
4.6	WASTE STREAMS		
4.6.1	Type of waste	Na_2SO_4 soln.	Chloride and fly ash slurry
	% water	85	95-97
4.6.2	Amount generated at full load	4 m^3/h	3.6 m^3/h
4.6.3	Disposal method	Removal of COD by acid and ozone.	By Hipsco
4.6.4	Disposal site and distance from plant, km	Discharged to sea with wastewater after treating.	By Hipsco
4.6.5	Potential effects on environment from sludge disposal	N.A.	Unknown
4.7	SLUDGE FIXATION PROCESS	N.A.	N.A.

(continued)

APPENDIX A (continued)

		MKK Nishinegoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
4.6	FGD EQUIPMENT DATA		
4.8.1	FIRST SCRUBBER		
4.8.1.1	Type	Sieve tray	Venturi
4.8.1.2	L/G[a] ratio, ℓ/Nm^3	0.6	0.6
4.8.1.3	Gas velocity through absorber or scrubber, m/s	1.8	38.2
4.8.1.4	Absorbent (adsorbent) flow, m^3/h		450 (11g)
4.8.1.5	Gas flow entering scrubber, Nm^3/h	N.A.	358,000
4.8.1.6	Gas temperature, °C	140	149
4.8.1.7	Gas composition		
	SO_2 in + out, ppm	1600 + 120	2270 + 2270
	% removal	92	0
	Particulate in + out, mg/Nm^3	50 + 35	700 + 200
	% removal	N.A.	70
	NO_x in + out, ppm	N.A. + 130	N.A.
	CO_2 in + out, %	N.A. + 12.1	12.01 + 12.01
	Other in + out	N.A.	N.A.
4.8.1.8	Material of construction	SUS	Carbon steel
4.8.1.9	Type of lining	N.A.	Acid brick
4.8.1.10	Absorbent and sand losses	N.A.	N.A.
4.8.1.11	Number of stages	3	1
4.8.1.12	Packing type and composition	N.A.	N.A.

[a] Liquid-to-gas ratio.

(continued)

APPENDIX A (continued)

		MKK Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
4.8.2	SECOND SCRUBBER	N.A.	
4.8.2.1	Type		Tower
4.8.2.2	L/G ratio, ℓ/Nm^3		0.44
4.8.2.3	Gas velocity through absorber or scrubber, m/s		2.99
4.8.2.4	Absorbent (adsorbent) flow, m^3/h		29.3
4.8.2.5	Gas flow entering scrubber, Nm^3/h		383,000
4.8.2.6	Gas temperature, °C		52
4.8.2.7	Gas composition		
	SO_2 in → out, ppm		2093 → 200
	% removal		90
	Particulate in → out, mg/Nm^3		200 → 100
	% removal		50
	NO_x in → out, ppm		N.A.
	% removal		N.A.
	CO_2		11.23% → 11.23%
	Other		0
4.8.2.8	Material of construction		Reinforced concrete shell with lining and 316LSS and alloy 20 internals.
4.8.2.9	Type of lining		Acid brick at bottom and tiles in absorbing section.
4.8.2.10	Number of stages		1
4.8.2.11	Packing type and composition		N.A.

(continued)

APPENDIX A (continued)

		MHI Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
4.8.3	CLEAR WATER TRAY		N.A.
4.8.3.1	Type	Packing tray	
4.8.3.2	L/G ratio, ℓ/Nm^3	0.2	
4.8.3.3	Source of water	Industrial water	
4.8.4	DEMISTER		
4.8.4.1	Type	Bending plate	Chevron
4.8.4.2	Number of passes (stages)	1	1
4.8.4.3	Material of construction	SUS	FRP
4.8.4.4	Method of cleaning	Water spray	Clean water sprays
4.8.4.5	Flow rate during cleanings, Nm^3/h	15	27.36
4.8.4.6	Frequency and duration of routine cleaning	Once daily	Very infrequent
4.8.5	REHEATER		
4.8.51.	Type	Direct heat (afterburner)	Gas fired direct heater
4.8.5.2	Heat load, kW	39,515	6700
4.8.5.3	Heat transfer surface area, m^2	N.A.	N.A.
4.8.5.4	Temperature at gas in, °C out, °C	60 200 (max.)	55 83
4.8.5.5	Reheater materials of construction	Refractory lining	316LSS

(continued)

APPENDIX A (continued).

		MHI Nishinagoya	DAVY POWERGAS, INC. D. N. Mitchell Plant
4.8.6	SCRUBBER TRAIN PRESSURE DROP, mbar		
4.8.6.1	First scrubber	39	9.9
4.8.6.2	Second scrubber	N.A.	30.0
4.8.6.3	Clear water tray	2	N/A
4.8.6.4	Demister	5	2.5
4.8.6.5	Reheater	7	2.5
4.8.6.6	Ductwork	N.A.	5.0
4.8.6.7	Total FGD system	54	50
4.8.6.8	I.D. fan	52	N.A.
4.9	SYSTEM MAKEUP WATER ADDED	31.7 m^3/h	21.6 m^3/h
4.9.1	Water added per mole of SO_2 removed	1 kg	N.A.
4.9.2	Source of makeup water	Industrial water	Lake water
4.9.3	Where is water added	At pump shaft to dissolve cake and scale	To venturi prescrubber
4.10	BYPASS SYSTEM		
4.10.1	Can flue gas be bypassed?	Yes	Yes
4.10.2	Height of gas discharge when bypassed, m	N.A.	Existing stack used
4.11	SLURRY DATA, pH, % solids by weight	N.A.	N.A.
4.11.1	Fresh slurry to scrubber		
4.11.2	Recirculating slurry		
4.11.3	Waste slurry		

(continued)

APPENDIX A (continued)

		MHI Nishinagoya	DAVY POWERGAS, INC. D. N. Mitchell Plant
4.12	SPENT SLURRY SYSTEM	N.A.	N.A.
4.12.1	Clarifiers		
4.12.1.1	Number		
4.12.1.2	Diameter, m		
4.12.2	Solid separation		
4.12.2.1	Type		
4.12.2.2	Cloth area, m^2		
4.12.2.3	Capacity (wet cake), tons/h		
4.13	SPACE REQUIRED, EXCLUSIVE OF SLUDGE DISPOSAL AREA, m^2	N.A.	N.A.
4.14	PERSONNEL, man-hours/yr		
4.14.1	Operating	N.A.	N.A.
4.14.1.1	Specialists	17,000 (2 persons)	26,000
4.14.1.2	Unskilled	17,000 (2 persons)	0
4.14.2	Maintenance	14,000 (7 persons)	18,000
4.15	ENVIRONMENTAL LOAD		3.6 m^3/h wastewater
4.15.1	Wastewater treatment process	Decomposition of COD by acid and ozone.	None (to fly ash pond)
4.15.2	Concentrations, %	N.A.	Fly ash, 5.0 Chlorides, Below 1 (vol.) Water 95 to 97
4.15.3	Appearance of flue gas discharge - any mist problems?	No	Steam plume (reheater is not yet in operation).

(continued)

APPENDIX A (continued)

		NEK Hishinegoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
4.16	FGD SYSTEM OPERATION		
4.16.1	% of boiler load FGD can operate	25-100	35-100
4.16.2	Lag time	None	N.A.
5.	ACTUAL COSTS FOR TOTAL FGD INSTALLATION, year	1972	1976
5.1	CAPITAL COSTS, $	N.A.	N.A.
5.1.1	Scrubber	N.A.	N.A.
5.1.2	Absorbent/regeneration system	N.A.	N.A.
5.1.3	Sludge disposal system	N.A.	N.A.
5.1.4	Site improvements	N.A.	N.A.
5.1.5	Land, roads, tracks, substation	N.A.	N.A.
5.1.6	Engineering costs	N.A.	N.A.
5.1.7	Contractor's fee	N.A.	N.A.
5.1.8	Interest during construction	N.A.	2,029,000
	Construction period	N.A.	N.A.
5.1.9	License fees	N.A.	N.A.
5.1.10	Other	N.A.	N.A.
5.1.11	Total, $	6 million (W 1.8 billion)	14,824,690 (installed)

W = Won

(continued)

APPENDIX A (continued)

		NKK Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
5.2	ANNUAL COSTS, $		
5.2.1	Interest	N.A.	N.A.
5.2.2	Depreciation	N.A.	N.A.
5.2.3	Insurance and taxes	N.A.	N.A.
5.2.4	Total fixed costs	1.5 million (includes maint. ¥ 450 million) 7 year depreciation	3,854,419
5.2.5	Labor and overhead Average man-hour cost cost/year	N.A.	
5.2.6	Raw material breakdown Unit cost Annual cost	N.A.	2,000,000
5.2.7	Utilities, unit cost, annual costs Electricity Water Steam Fuel	N.A.	
5.2.8	Maintenance cost/year	N.A.	513,000
5.2.9	Other costs, list	N.A.	Net operating - 2,151,440
5.2.10	Total variable cost/year	4.7 million (¥ 1.4 billion)	N.A.
5.2.11	Total annual cost	6.2 million (¥ 1.85 billion)	6,005,859
5.3	WASTE DISPOSAL COST, $ Unit cost Annual cost	N.A.	N.A.
5.4	BY-PRODUCT CREDIT, $	N.A.	361,560 total
5.4.1	By-product 1		32/mg
5.4.2	By-product 2		45/mg

(continued)

APPENDIX A (continued)

		MHI Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
5.5	NET ANNUAL COST, mills/kWh	N.A.	
5.5.1	Cost last year		8.1 Annual unit load factor 73% (92 MW/8000 h)
5.5.2	Cost a maximum continuous rate		N.A.
5.5.3	Cost without FGD		N.A.
6.	ENERGY REQUIREMENTS		
	Pumps	840 kWh	298 kW
	Fan	2320 kWh	N.A.
	Reheat	1.47 Mg/h (max)	6700 kW
	Reductant chemicals	N.A.	Natural gas - 0.034 kg/s
	Other	Power - 620 kW Steam - 18.2 Mg/h	3800 kPa steam - 7 kg/s
	Total	N.A.	N.A.
7.	PERFORMANCE HISTORY: OPERATING PROBLEMS AND SOLUTIONS		
7.1	Scrubber system	Trays and supports had to be reinforced because of cracks caused by vibrations.	Tray redesign required to achieve required turndown capability to 35% of design.
7.2	Demister	None	None
7.3	Reheater	None	Combustion air supply system had to be enlarged.

(continued)

APPENDIX A (continued)

	MHI Nishinagoya	DAVY POWERGAS, INC. D. H. Mitchell Plant
7.4 Fan	None	None
7.5 Sludge system	N.A.	N.A.
7.6 Regeneration system	Evaporator pluggage is being dealt with by washing more frequently and changing conditions.	None
7.7 By-product system	None	Dryer feed and recycle system redesigned.
7.8 Other	SO_2 cooler construction material was changed as a result of corrosion.	N.A.
7.9 Availability, %	99.5	97 (Over 62 days at above 90 SO_2 removal).

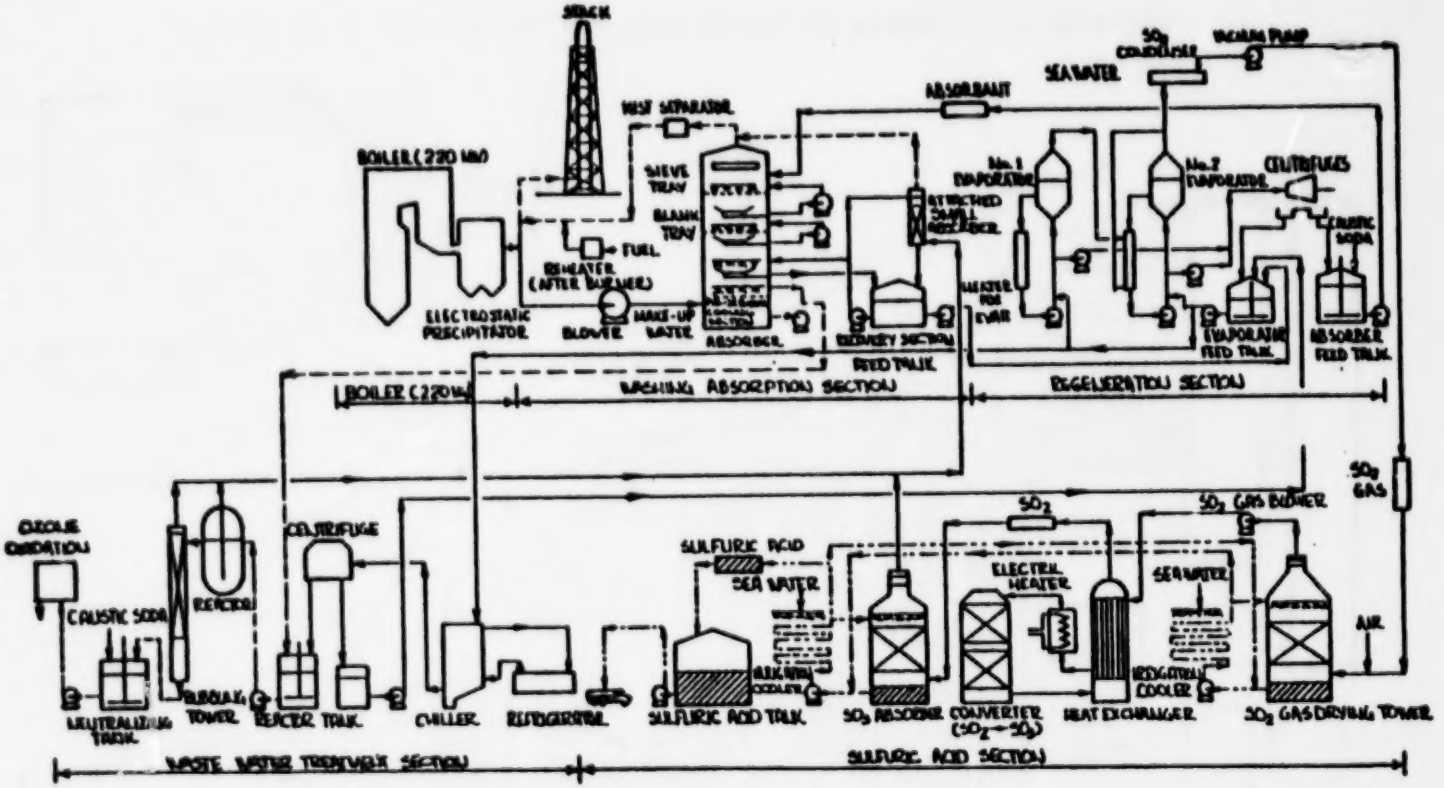

Figure A. Flowsheet of the Wellman-MKK process (Nishinagoya Station, Chubu Electric Power)

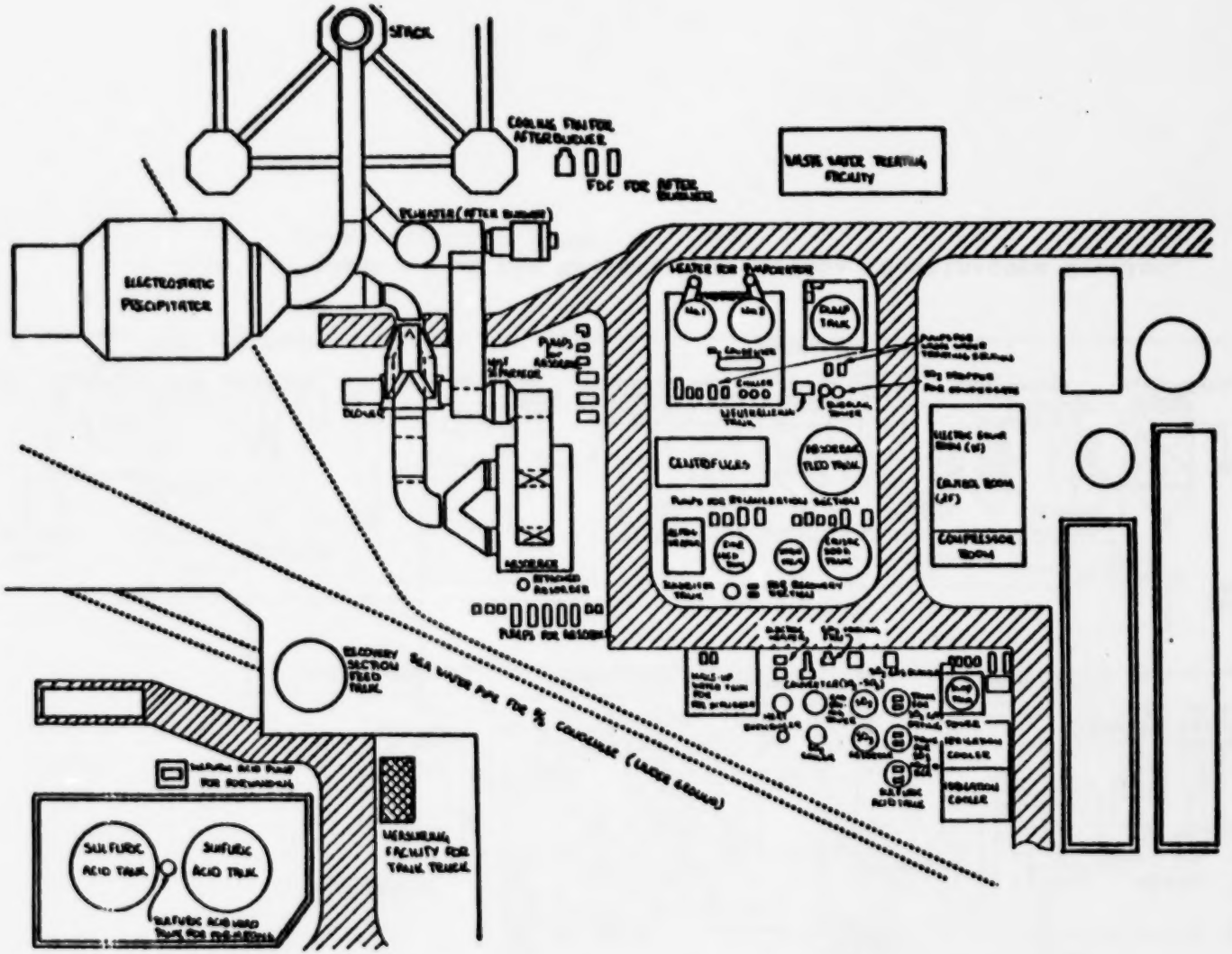

Figure B. Layout of the FGD plant at Nishinagoya Station

APPENDIX B

APPENDIX B

Availability Factor Computation

1. Divide boiler capacity by the number of modules and obtain MW/module = x

2. Multiply boiler capacity by number of hours during selected period (MW) = a

3. Add all down times due to module trouble for all modules during selected period = b

4. Add all down times due to boiler trouble or reduction in electricity demand for all modules during period = c

5. Availability factor = $\frac{[a - x\ (b + c)]\ 100}{a - x\ c}$ = %

APPENDIX C

APPENDIX C. CONVERSION FACTORS USED IN THIS STUDY

Multiply	By	To Obtain
ft	x 0.30480	m
ft^3	x 28.317	liter
Btu/lb	x 0.55554	kcal/kg
	x 2.32593	kJ/kg
Btu/h	x $0.29300 \cdot 10^{-3}$	kW
acfm	x 1.6992	m^3/h
$lb/10^6$ Btu	x 1.80005	mg/kcal
	x 0.42993	mg/kJ
gal/min	x 3.78541	liter/min
	x 0.22712	m^3/h
gal/1000 ft^3	x 0.13368	liter/m^3
ft^3/min	x 1.6992	m^3/h
kcal	x 4.1868	kJ
kcal	x $1.163 \cdot 10^{-3}$	kWh
scfm	x 1.60776	Nm^3/h
gr/ft^3	x 2.288	g/m^3
gr/scfm	x 2.4182	g/Nm^3
in. H_2O	x 0.2488	kPa
in. H_2O	x 2.54	mbar
gr	x 0.0648	g
ton (short)	x 1.9072	metric ton

1 MW corresponds to 3200 to 3500 m^3/h flue gas (coal)

1 scfm = 28.317 l (60°F / 14.7 psia)
(15.5°C / 760.28 torr)

26.796 l (0°C/760 torr) = 0.026796 Nm^3

6 SO_2 = 2.927 g/liter

MW = electric power generated unless other wise specified